

PB-217 473

PARTIAL DEMINERALIZATION OF UNCONTAMINATED BRACKISH WATERS

R. L. Sanks, et al

California University
Berkeley, California

May 1966

DISTRIBUTED BY:

NTIS

National Technical Information Service
U. S. DEPARTMENT OF COMMERCE
5285 Port Royal Road, Springfield Va. 22151

WP 00544

PB 217 473

LIBRARY
FWPLA, OHIO BASIN REGION
4876 Columbia Parkway
CINCINNATI, OHIO 45226

PARTIAL DEMINERALIZATION OF SALINE
SEWAGE BY ION EXCHANGE

PROGRESS REPORT NO. 2

PARTIAL DEMINERALIZATION OF UNCONTAMINATED BRACKISH WATERS

By

R. L. Sanks and W. J. Kaufman

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. Department of Commerce
Springfield VA 22151

May 1966

Sanitary Engineering Research Laboratory
College of Engineering
and
School of Public Health
University of California
Berkeley

INTRODUCTION

The continuing increase in water demand in the United States, coupled with a dwindling reserve of uncommitted fresh water sources, has prompted extensive efforts to find an economical means of producing drinking water from the inexhaustible seas. A significant effort has also been directed toward the reuse of waste waters and the demineralization of brackish waters at the margin of mineral acceptability. Brackish water is a particularly valuable potential water resource for many inland communities because it is readily available and can be developed on a small scale, often with only a modest capital investment. Brackish ground waters containing 1000 to 3000 mg/l of dissolved solids underlie about 50 per cent of the land area of the United States, and many of the surface waters in the states of the southwest and on the eastern slopes of the Rocky Mountains are brackish. In many localities mineral buildup due to consumptive use of near-brackish waters by irrigation is increasing the number of supplies that are unsuitable for direct domestic use.

It is evident that municipal need for methods of demineralizing brackish waters or for reducing the mineral content of marginal waters must increase in the future. Moreover, success in reducing the cost of sea water conversion will have little bearing on the resolution of brackish water supply problems of small communities at great distances from the sea. Brackish ground waters have many of the desirable attributes of other groundwaters: freedom from particulates, minimal probability of pathogenic contamination, availability at point of use, and natural storage.

Ion exchange systems are especially well suited to processing fresh waters and are, of course, widely employed for this purpose by industry. There are more than 17,000 ion exchange plants in the United States producing very low-mineral boiler feed or industrial process water. Most of these plants are semi or fully automatic and operate with a minimum of attention. However, the application of ion exchange to the demineralization of brackish water for municipal supply has received but little attention, although considerable effort has been focused upon electrodialysis. The primary objection to ion exchange has been its cost in comparison with other solutions to the problem. It is believed that the diminution of alternate sources of water for municipalities, coupled with the possibility of more economical ion exchange systems, warrants a review of the potential of ion exchange.

Objectives and Scope of the Study

The broad objective of the study was to optimize the design and operating cycle of ion exchange systems for the specific purpose of demineralizing brackish waters for municipal use. As regenerant chemicals comprise a major item of cost in ion exchange demineralization, special emphasis was placed on ascertaining the most efficient exhaustion-regeneration cycle. Throughout the study advantage was gained by using the generally lower domestic criteria of mineral acceptability in comparison with the usual industrial requirements for low conductivity water.

In order to identify and measure the pertinent technological factors, laboratory investigations were made using the strong acid cation-weak base anion exchange resin system with conventional regenerants. Both types of exchangers were investigated from three standpoints: (1) a preliminary study using a local well water having a total dissolved solids concentration

of 770 mg/l to ascertain breakthrough concentration histories, exchange capacities, selectivity coefficients and to confirm certain questions regarding exhaustion and regeneration rates and directions and the selection of regenerants; (2) a study of cyclic operation to establish exchange efficiencies under simulated normal conditions as determined by level of regeneration and regenerant reuse; and (3) a pilot plant study using a water having a total dissolved solids concentration of 2000 mg/l with operating conditions selected to conform to those attainable in municipal practice. In the final phase of the study a detailed analysis was made to establish the relation of process economics to the concentration of the normal mineral constituents of brackish ground waters.

Previous Work

In 1944 Bachmann [1] described the demineralization of a well water containing 19 me/l NaHCO_3 and 10 me/l NaCl . Half of the water was treated with a strong acid resin (Amberlite IR-1) which reduced the sodium to 0.8 me/l. After blending with by-passed water and degasifying, the plant effluent contained 15 me/l sodium, 5 me/l bicarbonate, and 10 me/l chloride. By three-stage regeneration, in which spent acid was reused twice, the overall exchange efficiency (i.e., cations removed from the net product per unit of regenerant) was 80 per cent. The total cost was \$0.165/1000 gal at a plant capacity of 0.6 mgd. Using 60°Bé sulfuric acid at \$13.00/ton, chemical costs were 41 per cent of the total cost.

The advantage of counterflow regeneration was reported by Spiegler, Juda, and Carron [2] in 1952. They obtained an overall exchange efficiency of 70 per cent in a two-stage (cation-anion-cation-anion) pilot plant by reusing spent acid in downflow regeneration and exhausting upflow with brackish water. The exchanger was a strong-acid Dowex-30 resin. The

overall exchange efficiency for the weak-base Amberlite IR-4B exchanger was slightly over 67 per cent. Both efficiencies required a strict regimen of conservation of rinse and wash waters. The cost to remove 1000 mg/l NaCl was estimated to be \$0.60/1000 gal.

In 1956 Yoseph [3] reported on a fully automatic demineralization plant that replaced a faulty thermocompression distillation plant on Matagorda Island. The plant consisted of a sodium zeolite softener followed by cation and anion exchangers. It reduced the mineral concentration from 2100 mg/l to 50 mg/l at a chemical cost of \$2.04/1000 gal. The plant operated 12 hours per day and produced 3600 gallons of usable product. Including the amortization of the \$27,400 investment in equipment and neglecting labor, the cost per 1000 gallons exceeded \$4.00.

Kunin and Vassiliou [4] have demonstrated the superiority of weak electrolyte resins with respect to regeneration efficiency and rinse water requirements. They have also proposed [5] a counterflow anion-cation-anion system in which weak electrolyte exchangers may be used to deionize brackish waters up to 3000 me/l TDS with little leakage and with high regeneration efficiency. In a 12-month pilot plant study of this system in Italy, Sturla [6] concluded that for the purpose of producing low solids water suitable for industrial purposes, chemical costs were much less than for conventional systems operated at normal regeneration levels. Unfortunately, no cost data were presented for the Kunin system. In the SUL-biSUL process announced in 1965 by the Nalco Chemical Company [7] cation removal is accomplished by a conventional strong-acid exchanger regenerated with sulfuric acid while anions are removed in a strong-base exchanger regenerated to the sulfate form by raw water. During exhaustion the anion bed is converted to the bisulfate form by the acidic

influent, thus making available 50 per cent of the exchange sites for the sorption of other anions. The effective capacity of the anion bed is limited to half of the normal capacity and hence a larger reaction vessel is required. The regeneration rate of the anion exchanger is low, although it can be increased to a satisfactory value by the addition of lime at about twice the stoichiometric amount needed to convert HSO_4^- sites to SO_4^{2-} sites.

LABORATORY INVESTIGATION

A general view of the ion exchange plants employed in this investigation is shown in Figure 1. The components are identified as follows: A, pumps; B, pressure release manometer; C, effluent pH electrode; D, adjustable plug and water seal; E, flow meter; F, double shut-off quick-connection; G, control valves to produce upflow, downflow, or bypass; H, interval timer for fraction collector; I, automatic 6-point pH switch; J, pH meter-amplifier; K, pH recorder. The conductivity cells are behind the panels.

Feedwater sufficient for a series of runs was stored in 55-gallon stainless steel drums while the acid decationized water was held in polyethylene barrels until ready for use. In order to minimize the backwashing water requirements, feedwater was circulated through a 5-micron Cuno Micro-Klean filter and passed downward through the resin beds by Sigmamotor pumps powered by Zeromax units or by Bodine NSH-5⁴ DC motors with Minarik SH-5⁴ speed controls. The resin beds were 1 or 1-1/2 inches in diameter and from 24 to 36 inches in height. Each of the four columns was equipped with adjustable plugs and water seals that could be raised or lowered as desired so as to minimize the system void volume. The resins were supported by glass beads and confined at the

upper water seals by fiberglass pads. The influent could be halted automatically at any desired leakage concentration by sensors responding to either conductivity or pH, and the latter index of performance was continuously recorded. Carbon dioxide in the cation exchanger effluent was partially removed by aeration.

The feedwater for the preliminary and cyclic phases of the study was obtained from a well adjacent to the laboratory and was selected because of its availability, its relatively high mineral content (770 mg/l TDS), and its representative composition. For the final phase of the investigation, the pilot plant study, a synthetic water was prepared having a TDS of 2000 mg/l and a relative distribution of mineral constituents nearly equal to that of the well water. The concentrations of the major ions present in these two waters are shown in Table I.

TABLE I
FEEDWATER ANALYSES

Constituents	Well Water			Synthetic Water		
	Representative	Max.	Min.	Representative	Max.	Min.
Na + K, me/l	3.56	3.68	3.54	9.53	9.53	8.04
Mg, me/l	4.70	4.72	4.56	12.47	12.91	12.47
Ca, me/l	<u>3.30</u>	3.30	3.28	<u>8.75</u>	8.97	8.75
Sum, me/l	11.56			30.75		
HCO ₃ , me/l	4.56	4.56	4.20	12.15	12.15	11.27
Cl, me/l	3.30	3.34	3.30	8.90	8.94	8.90
NO ₃ , me/l	0.41	0.41	0.41	1.15	1.15	1.15
SO ₄ , me/l	<u>3.02</u>	3.02	2.94	<u>8.16</u>	8.20	8.16
Sum, me/l	11.29			30.36		
Avg. TDS, computed, mg/l	770			2066		
Avg. Conductivity, micromhos/cm	1100			2500		

Chemical Determinations

Sodium and potassium were determined by flame spectrophotometry. Calcium was measured by titration with EDTA using hydroxy naphthol blue as the indicator. Magnesium was determined as the difference between calcium and hardness, which was determined by EDTA with a mixture of eriochrome black T and methyl red to the bright green endpoint. Bicarbonate, chloride, and sulfate were measured by titration, mercurimetric, and turbidimetric methods respectively in accordance with Standard Methods [8]. Nitrate was determined by the brucine method modified by Jenkins and Medsker [9]. Carbon dioxide in the highly acid cation exchanger effluent was determined by placing aliquots in an excess of NaOH and titrating from pH 8.3 to 4.3.

Selection of Exchange System and Resins

The costs of the various ion exchange systems in common use were discussed by Kahler and Reents [10] who concluded that the strong acid cation-weak base anion exchange system is by far the least expensive. Carbon dioxide is not removed, however, and if it is objectionable it can be reduced to about 5 ppm by aeration. The strong acid-aeration-weak base ion exchange system was selected for the laboratory investigations.

Three strong acid resins, Duolite C-20 and C-25 and Amberlite 200 were employed. Duolite C-20 is a sulfonated polystyrene resin with a gel or solid structure containing 8 per cent divinylbenzene (DVB). Duolite C-25 is similar except that it is porous and normally contains 5.5 per cent DVB. Amberlite 200 is a macroreticular sulfonated polystyrene resin with 20 per cent DVB. It is physically and chemically durable and its macroreticular structure of small pores averaging $300 \text{ \AA}$ in diameter makes it more resistant to organic fouling. In the cyclic operations, Duolite C-25 was selected in

preference to C-20 because of its lower rinse requirement and its pre-supposed resistance to fouling. Both C-25 and Amberlite 200 were used in the pilot plant investigations.

The weak base anion exchange resin selected was Amberlite IRA-93. It is a polystyrene DVB resin with functional groups or exchange sites consisting only of tertiary amine. Both matrix and functional groups are resistant to chemical degradation and it has been advocated for use with aggressive waters. Its macroreticular structure with pores averaging 1400 Å in diameter permits organic ions to be eluted.

CATION EXCHANGE

Preliminary Investigation

Although the ion exchange literature is extensive, optimum operating conditions for exhaustion, regeneration, and rinse for the production of potable water from brackish water are not available and it was necessary to determine them experimentally. The preliminary phase of the investigation consisted of a series of individual runs made with well water. The operating conditions were selected to approximate the optimum exhaustion-regeneration cycle. At the beginning of each exhaustion run, the resin was completely regenerated. Before each regeneration run, the resin was exhausted to 60 per cent sodium breakthrough as this was judged to be a reasonable estimate of the desirable maximum leakage rate.

Both C-20 and C-25 resins were exhausted to complete breakthrough of all cations in order to determine resin exchange capacity, equilibrium parameters and the relation between pH and breakthrough. As shown in Figure 2, which is typical, the breakthrough curve for sodium was very abrupt and was inversely related to that for hydrogen ion. Halting the

feed automatically at pH 6.0 provided accurate control of throughput for conditioning the resin bed for regeneration runs.

Sixteen regeneration runs were made beginning with resins in the partially exhausted state. Acid concentration, kind of acid, flow rate, and flow direction were the variables studied. Complete concentration histories, as exemplified by Figure 3 were obtained, permitting the influence of the amount of regenerant (i.e., regeneration level) on exchange efficiency to be estimated and confirmed in subsequent cyclic operations. It was expected that the exchange efficiency in optimum cyclic operation would be about 80 per cent, so that comparisons of efficiency were made at a regeneration level of $T = 1.25$, where T is the ratio of ion throughput to the stoichiometric throughput corresponding to the exchange capacity of the resin. As shown in Figure 3, the rate of removal of cations is high at the early stages of regeneration, decreasing rapidly. The exchange efficiency can be obtained by dividing the sum of the areas under the concentration histories of each metallic cation by the total throughput of hydrogen ion. Allowance must be made for the lag due to the system void volume and for the gradual decrease in concentration during rinse. It is evident that exchange efficiency increases as the regeneration level decreases.

As expected, the exchange efficiency improved with increasing acid concentrations as shown in Figure 4. But unexpectedly, as shown in Figure 5, the efficiency was nearly independent of the flow rate for the range of flow rates studied. At the highest concentration of acid, the efficiency actually increased slightly as the rate of flow increased. This is fortunate for two reasons. The regeneration and cycle times are reduced, allowing a smaller resin bed to be used. The high regeneration

rate also allows the removal of calcium sulfate from the column before its precipitation occurs. The comparison of concurrent and countercurrent regeneration clearly demonstrated the superior efficiency of the latter.

Cyclic Operations

Three series of cyclic studies were made with C-25 resin. In each series exhaustion and regeneration cycles were continued until their individual concentration histories became essentially identical, i.e., until cyclic stability was reached. The purpose was to determine the exchange efficiencies that would be sustained at three regeneration levels and with acid reuse. The resin was exhausted at a downflow rate of 3.5 gpm/cu ft of resin. The flow was stopped automatically when the pH reached 6.0. The resin was then backwashed with one bed volume of distilled water at a rate of 0.8 gpm/cu ft which produced a bed expansion of about 50 per cent.

The concentration histories in Figure 6 show the benefit of recovering once-used acid and its reuse in two preregeneration steps. Acid recovery was initiated in the first cycle by passing upward through the column 153 ml of 1.05 N sulfuric acid. After passage of the system fluid volume, two 76-ml effluent fractions were collected. After 20 hours for the precipitation of calcium sulfate, the fractions were filtered, the first fraction was diluted to 3 times its original volume, and the second fraction diluted to twice its original volume. In the next regeneration cycle the first diluted fraction was passed through the column followed by the second and the corresponding effluent volume of twice-used acid, 382 ml, was wasted. The column then received 153 ml of new acid and, after allowing for the system fluid volume, two portions, each containing 76 ml of partially spent acid, were collected as before.

The column was then rinsed upflow with 2.5 bed volumes of distilled water and the next cycle commenced. In Figure 6 the decrease in Ca and Mg concentrations at throughput volumes of 0.50 to 0.65 liters and their rapid increase at 0.70 liters corresponds to the transition from pre-regeneration to regeneration with new acid. The area under the effluent hydrogen ion curve is small before a throughput of 0.75 liters, indicating efficient use of the preregenerant. The exchange efficiency was calculated as the metallic cations removed from the total volume of product water divided by the hydrogen ions supplied as new acid. It was verified by a material balance on hydrogen ion throughput, i.e., $(H_{in} - H_{waste})/H_{in}$. Figure 6 represents the tenth regeneration cycle of the series at 1.26 lb H_2SO_4 /cu ft and is typical of all cycles after the sixth. The exchange efficiency was 91 per cent in this cycle.

The average results of the last four of the ten cycles in each series are shown in Figure 7. As they should be, the exchange efficiencies are higher than those calculated from manufacturer's data because the latter did not reuse acid, exhaust nearly to complete breakthrough, or regenerate countercurrently. These operational differences also account for the differences in leakage. The exchange efficiencies do not include loss of product water for rinsing, acid dilution, or backwash. Correcting the highest efficiency obtained, 91.7 per cent, to a net product basis results in a net operating efficiency of about 87 per cent for the 770 mg/l feedwater studied. The difference between exchange efficiency and net operating efficiency increases with feedwater mineral content, reaching about 10 per cent at a TDS of 2000 mg/l.

Pilot Plant Operations

The purpose of the pilot plant operations was to determine the operating exchange efficiencies that could be maintained under prototype operating conditions at the optimum regeneration level established by cyclic operation for a moderately brackish water. The only water used for feed, acid dilution, or rinse was the 30 me/l synthetic water (2000 mg/l TDS; cf. Table I) or its derivative obtained from the cation columns. The decationized water was contaminated by the leakage associated with the high exchange efficiencies and contained from 110 to 160 mg/l of metallic cations and from 12.5 to 15.5 me/l of hydrogen ions.

Two series of runs were made on each of two resins, Duolite C-25 and Amberlite 200. The operations were similar for all runs in both series, except that in one series acid was not reused while in the other acid was stored, filtered, and reused as described for the cyclic study. In each series product water was conserved by several means: (1) rinse water was used for acid make-up; (2) backwashing was eliminated (in a prototype, backwashing might be necessary twice a day which would occur only once in 12 or more cycles); (3) at the beginning of regeneration a large portion (from 69 to 74 per cent) of the system fluid volume of water was assumed to be returnable to feed and need not, therefore, be wasted; and (4) the freeboard above the top of the resin bed was kept at a minimum to reduce the system fluid volume and longitudinal mixing in the system.

Exhaustion was carried out downflow at a rate of 3.5 gpm/cu ft and automatically halted at pH 4.0. This permitted a total cation leakage of 17 to 23 per cent (on a me/l basis) of the feed concentration, depending upon the resin used and whether acid was reused.

Regeneration was upflow at a rate of 2 gpm/cu ft using 1.25 lb H_2SO_4 /cu ft per cycle. Tests showed this was equivalent to 1.35 lb commercial 66°Bé acid/cu ft. Regeneration was followed by a rinse of cation exchange product water until the conductivity decreased to 6000 micromhos/cm, at which point the cation concentration in the effluent was nearly equal to the concentration of cations in the feed. In a second series of tests, regeneration was preceded by a preregeneration step similar to the preregeneration described for the cyclic phase of this investigation except that the first fraction of reclaimed acid was diluted to 2.33 times its original volume with rinse water and the second fraction was not diluted at all.

The runs in each series were continued until the total variation in ions removed per cycle from the throughput water did not exceed 3 per cent. The average results from the last four runs of each series are given in Table II. The benefit of acid reuse is shown by an increase of 13 per cent in efficiency, by a 9 to 15 per cent reduction in leakage, and by a 10 per cent increase in throughput. The reuse of acid decreased the quantity of waste water from the C-25 exchanger, but not from the Amberlite 200 exchanger. Comparison of the performance of the resin showed C-25 to be slightly superior to Amberlite 200 with respect to efficiency, leakage, and waste water.

Operating characteristics for a prototype exchanger decationizing a feed of 2000 mg/l of dissolved solids are given in Table III. Figure 8 illustrates the operating cycle for the C-25 resin with acid reuse. The data shown were drawn from Table III. Backwashing is not included in the table and figure because it is likely to be necessary only about twice daily or once in 10 to 15 cycles. When required, the column can be

TABLE II
RESULTS OF CATION EXCHANGE IN PILOT PLANT OPERATIONS

Characteristic	Acid Used Once		Acid Reused	
	C-25 ^a	200 ^a	C-25 ^a	200 ^a
Feedwater cation conc., me/l	29.9	29.9	30.7	30.7
Effluent cation conc., avg.				
Na, me/l	5.9	6.5	4.8	6.0
Mg, me/l	0.2	0.6	0.3	0.3
Ca, me/l	<u>0.0</u>	<u>0.0</u>	<u>0.1</u>	<u>0.1</u>
Total, me/l	6.1	7.0	5.2	6.4
Total feedwater volume, liter	14.20	14.40	15.30	15.90
Concentrated acid, liter	0.01	0.01	0.01	0.01
Unrecoverable waste, liter	1.23	1.50	0.94	1.62
Recoverable waste, liter	<u>0.60</u>	<u>0.60</u>	<u>0.60</u>	<u>0.60</u>
Net product per cycle, liter	12.38	12.31	13.77	13.69
Decationized water used, liter	1.83	2.10	1.54	2.22
Acid dilution, liter	0.00	0.00	0.02	0.00
Rinse, liter	1.83	2.10	1.52	2.22
Regeneration volume, liter	0.40	0.40	1.06	1.06
Cations removed per cycle				
Total feedwater, me	338	330	390	387
Net product, me	294	282	351	333
Concentrated acid used, me	420	420	419	426
Exchange efficiency, %	80	79	93	91
Net (operating) efficiency, % (corrected to net product)	70	67	84	78

^aVolume resin = 1030 ml, column height = 90 cm, and system fluid volume = 876 ml.

backwashed with deionized product water which can largely be recovered by filtration. The waste volumes are shown to be very low (from 7 to 12 per cent of the product), although this was not the objective of the study. If disposal of waste were a problem, the volume of waste could be reduced considerably with insignificant changes in efficiency. The cycle times are very short and fully automatic operation is required. Fortunately, the cost of fully automatic operation is not significantly greater than semi-automatic operation with manual initiation.

TABLE III
OPTIMUM OPERATING CHARACTERISTICS FOR CATION
EXCHANGE AND 2000 mg/l FEEDWATER

Characteristic Per Cu Ft of Resin for One Cycle	Acid Used Once		Acid Reused	
	C-25	200	C-25	200
Net product, gal	90.0	89.4	100.0	99.4
Recoverable waste, gal	4.4	4.4	4.4	4.4
Unrecoverable waste, gal	<u>9.0</u>	<u>10.9</u>	<u>6.8</u>	<u>11.8</u>
	103.4	104.7	111.2	115.6
Concentrated acid ^a , gal	<u>0.1</u>	<u>0.1</u>	<u>0.1</u>	<u>0.1</u>
Total feedwater, gal	103.3	104.6	111.1	115.5
Decationized rinse, gal	13.3	15.2	11.0	16.1
Decationized water for acid dilution, gal	<u>0.0</u>	<u>0.0</u>	<u>0.1</u>	<u>0.0</u>
Operation water, gal	13.3	15.2	11.1	16.1
Cycle time, min				
Exhaustion at 3.5 gpm	29.5	29.9	31.8	33.0
Regeneration at 2 gpm	1.5	1.5	3.9	3.9
Rinse at 2 gpm	<u>6.7</u>	<u>7.6</u>	<u>5.5</u>	<u>8.1</u>
Total, min	37.7	39.0	41.2	45.0
Operating exchange capacity, Kgr (based on total feedwater)	7.2	7.0	8.3	8.2

^aRegeneration level: 1.25 lb H₂SO₄ per cu ft. Equivalent to 0.088 gal of 66°Bé sulfuric acid.

ANION EXCHANGE

Preliminary Investigation

The preliminary phase of the anion exchange investigation consisted of a series of 20 runs using decationized well water and dilute hydrochloric and sulfuric acid feeds. The 770 mg/l well water was completely decationized by passage through two strong acid resin beds in series after which it was degasified to less than 70 mg/l of CO₂. The acids were reagent grade diluted to 0.1 N or to 0.01 N with distilled water. Prior to the beginning of each exhaustion run, the resin was fully regenerated and each regeneration run was preceded by exhaustion of the resin to a particular degree of breakthrough at which the flow was stopped automatically by the conductivity meter.

An exhaustion run to complete equilibrium was made with decationized well water in order to determine resin exchange capacities, equilibrium parameters, the characteristics of the breakthrough curves and the relation between them and pH and conductivity. The concentration histories are shown in Figure 9. Chloride was the first ion to emerge followed by nitrate and sulfate. Conductivity was an excellent indicator of breakthrough, while pH varied only slightly at a chloride breakthrough of 50 per cent and was therefore not a satisfactory indicator. The effect of feed concentration and flow rate on breakthrough was studied using 0.01 N and 0.1 N hydrochloric acid. At breakthrough concentrations of less than 25 per cent ($C/C_0 < 0.25$) there was little effect, either of feed concentration or flow rate. At higher breakthroughs an increase in concentration or flow rate resulted in increased operating exchange capacities. The total exchange capacity of weak base resins varies with the conditions of operation. The total exchange capacity determined from

Figure 9 was 4.68 me/g dry OH form resin. The recommended exchange capacity determination [11] yielded an average of 4.81 me/g, but both higher and lower values were found in other runs.

The effect of regenerant flow direction, rate, and chemical was studied by determining exchange efficiencies or, more properly, regeneration efficiencies.* The fully regenerated resin was exhausted with decationized well water (or with sulfuric or hydrochloric acid for some determinations) to a breakthrough of $C/C_0 = 0.3$. The quantity of anions sorbed by the resin was determined by a material balance. In one example this was 370 me. Without backwashing, a stoichiometric amount (370 me) of regenerant was passed through the column followed by distilled rinse water. The regeneration efficiency was found by comparing the area under the breakthrough curve of OH ion with the input OH ion and verified by comparing the sum of the areas under the other breakthrough curves with the input OH ion. Figure 10 depicts a countercurrent regeneration with sodium hydroxide in which the regeneration efficiency was 99.9 per cent. Concurrent regeneration resulted in a 2 to 5 per cent reduction in efficiency. Flow rates below 1 gpm/cu ft did not alter the efficiency, but at 2 gpm/cu ft there was a reduction in efficiency. The reduction was 7 per cent when the resin was exhausted with sulfuric acid, but the reduction was less than 2 per cent when the resin was exhausted with hydrochloric acid.

Sodium hydroxide, ammonia, and sodium carbonate regenerants applied concurrently in stoichiometric amounts resulted in efficiencies of 96, 93, and 80 per cent, respectively. Regeneration was followed by exhaustion

*Regeneration efficiency represents the amount of regenerant sorbed divided by the amount applied. It is only indicative of column performance. Regeneration efficiency equals exchange efficiency when measured at cyclic equilibrium.

with decationized well water to determine the suitability of the product water. Following regeneration with sodium hydroxide and ammonia, the product was satisfactory and the water was palatable. Only a trace (less than 0.01 me/l) of ammonia was found in the product water following ammonia regeneration. However, sodium carbonate regeneration yielded a poor quality effluent with low pH. Moreover, during regeneration, CO_2 gas was evolved which disrupted the column. On an overall cost-efficiency basis, ammonia is superior to sodium hydroxide and sodium carbonate. Ammonia is often used to recondition weak base resins regenerated with caustic and thus its use as a regenerant is beneficial to resin performance.

Decationized water, deionized water, and raw water were compared for use as rinse water in three runs. Decationized water was slightly more satisfactory than raw water and deionized water. Raw water caused $\text{Mg}(\text{OH})_2$ to form at the top of the column. It could be removed only with difficulty, and its formation precludes the use of raw water for rinse. The change of pH during rinsing was slight, but there was a great change in conductivity which provided accurate quality control for rinsing.

Cyclic Operations

Cyclic operations using 0.025 N sulfuric acid as feedwater and sodium hydroxide as regenerant were made to determine the effect of extent of breakthrough upon exchange efficiency, yield, and leakage. In one series, the exhaustion was halted automatically at a $\text{SO}_4 + \text{Cl} + \text{NO}_3$ breakthrough of 5 per cent ($\text{C}/\text{C}_0 = 0.05$) and in the other series exhaustion continued to $\text{C}/\text{C}_0 = 0.77$, which was the highest ratio that could be accurately controlled by the conductivity meter. At the beginning of each series, the resin was conditioned by complete regeneration and thorough rinsing. The quantity of anions sorbed by the resin in the first

exhaustion run was determined to be 368 me for the 5 per cent breakthrough series and 428 me for the 77 per cent breakthrough series. In each series the resin was regenerated with a stoichiometric amount (i.e., 368 or 428 me) of 1 N sodium hydroxide. The exchange efficiency was computed as the quantity of OH ions sorbed by the resin divided by the OH ions supplied by the regenerant (cf. footnote, p. 17).

For the 5 per cent breakthrough series the average exchange efficiency was 96 per cent and for the 77 per cent breakthrough series it was 95 per cent. The average leakage of sulfate in the product water was 0.1 per cent in the 5 per cent breakthrough series, but was 10.8 per cent in the 77 per cent breakthrough series and the pH was low. Clearly there is no advantage in allowing breakthrough to high ratios. The extent of breakthrough can be used as a means of controlling the pH of the product water with an insignificant effect upon efficiency.

Pilot Plant Operations

The purpose of the pilot plant operations was to determine exchange efficiencies that could be maintained under prototype operating conditions. The feed, regenerant dilution water, and rinse water were the effluent from the cation exchange columns and thus contained the cations shown in Table II in addition to the chloride, sulfate, and nitrate given for the 2000 mg/l synthetic water in Table I. The bicarbonate in the feed was reduced to less than 2 me/l by aeration. Bicarbonate was not removed by the weak base resin and its presence in the feed did not significantly influence the performance of the resin.

Exhaustion was carried out downflow at a rate of 3 gpm/cu ft and halted automatically at a conductivity of 2000 micromhos/cm which corresponds to a breakthrough of nearly 50 per cent of the strong acid anions.

At this breakthrough the pH of the effluent was about 2, but the pH of the combined effluent exceeded 7.0. The resin was not backwashed.

Two anion exchange columns were operated in an identical fashion, except that one was regenerated with sodium hydroxide and the other was regenerated with an equivalent amount of 30 per cent aqueous ammonia stored in a collapsible bag to protect it from air and prevent the loss of ammonia gas. Regeneration was upflow at a rate of 1 gpm/cu ft and at a regeneration level of 0.944 me/ml of resin in its irreversibly swollen OH form. This corresponds to 2.35 lb NaOH/cu ft and to 1.00 lb NH₃/cu ft. The amount of base actually used exceeded the values given by approximately one per cent so as to neutralize the mineral acidity. This level of regeneration corresponded to about 80 per cent of the total exchange capacity and ensured a layer of exhausted resin at the top of the bed to sorb any leakage of OH ion during regeneration, thus promoting high efficiency. As shown by Figure 11, there was no leakage of OH ions. Any inefficiency of operation is therefore attributable to the need for decationized water for base dilution and for the rinse which immediately followed regeneration. The rinse was stopped automatically at a conductivity of 2000 micromhos/cm. In Figure 11, which was the last run of the series, excess rinse water was used to obtain more complete concentration histories.

Each series of runs was continued until the total variation in ions removed per cycle from the feed did not exceed 2 per cent. The average results from the last four runs of each series are given in Table IV. Operating exchange efficiencies were calculated by dividing strong acid anions (Cl + NO₃ + SO₄) removed from the net product by the gross amount (324 me) of base used in regeneration. The table shows that

TABLE IV
RESULTS OF ANION EXCHANGE IN PILOT PLANT OPERATIONS

Characteristic ^a	Regenerant	
	NaOH	NH ₄ OH
Feedwater anion concentration less HCO ₃ , mg/l	18.29	18.21
Effluent cation concentration, avg.		
Cl, mg/l	4.12	5.48
NO ₃ , mg/l	0.40	0.56
SO ₄ , mg/l	0.24	0.34
HCO ₃ , mg/l	<u>0.66</u>	<u>0.62</u>
Total less HCO ₃ , mg/l	4.76	6.38
Total feedwater (= net product) volume, liter	21.30	23.40
Regenerant		
Concentrated base, liter	0.02	0.02
Decationized dilution water, liter	<u>0.29</u>	<u>0.28</u>
Total	0.31	0.30
Rinse (decationized water), liter	0.70	0.68
Waste volume, liter		
Unrecoverable waste	0.76	0.73
Recoverable waste	<u>0.25</u>	<u>0.25</u>
Total	1.01	0.98
Anions removed per cycle, me	288	277
Concentrated base used, me	324	324
Net (operating) exchange efficiency, %	89	86

^aVolume resin = 339 ml in swollen OH form, column height = 67 cm, and system fluid volume = 422 ml.

the exchange efficiency is a little less for ammonia than for sodium hydroxide regeneration and the leakage for ammonia regeneration is a little higher. The leakage could be reduced, but only at the expense of a greater reduction in efficiency. Because ammonia regeneration costs about 60 per cent of caustic regeneration, it is, however, to be preferred.

Operating characteristics for a prototype anion exchanger are given in Table V. Backwashing is not included for the same reasons given for omitting it in cation exchange. The waste volumes are shown to be only about 3 per cent of the product. It would be possible to reduce this volume slightly by stopping the rinse sooner or by using part of the waste rinse for dilution of the base. The times required for the various parts of the exhaustion-regeneration-rinse cycle do not correspond to those for the cation exchanger. Hence a direct connection between cation and anion exchangers (which is the rule in industrial applications) would result in compromises that would either decrease efficiency or create idle time and consequently larger resin beds. A degasifier between the cation and anion columns is advantageous because CO_2 is readily removed from acid solutions and the storage normally associated with degasifiers permits each exchanger to operate at peak efficiency virtually independent of the other exchanger. The water returnable to feed is the system water stored in the resin bed after due allowance for longitudinal mixing with the regenerant as it displaces the pore water.

TABLE V

OPTIMUM OPERATING CHARACTERISTICS FOR ANION EXCHANGE
AND 2000 mg/l FEEDWATER DECATIONIZED

Characteristic Per Cu Ft of Resin (Swollen OH Form) for One Cycle	Regenerant	
	NaOH	NH ₄ OH
Net product, gal	470.0	516.0
Total waste, gal	22.2	21.6
	492.2	537.6
Concentrated aqueous base, gal	0.4	0.5
Total input water*, gal	491.8	537.1
Recoverable waste, gal	5.5	5.5
Net input water*, gal	486.3	531.6
Decationized rinse*, gal	15.4	15.0
Decationized water* for base dilution, gal	6.4	6.2
Operation water*, gal	21.8	21.2
Base**, lb (NaOH-50% aqueous; NH ₃ -30% aqueous)	2.36	1.01
Cycle time, min		
Exhaustion at 3.5 gpm	134.1	147.3
Regeneration at 1 gpm	6.8	6.6
Rinse at 1 gpm	15.4	15.0
Total, min	156.3	168.9
Operating exchange capacity, Kgr	18.6	17.9

* Product water from cation exchanger (decationized).

** Gross pounds as NaOH or NH₃.

ECONOMIC ANALYSIS

In order to increase the significance of comparisons of cost between ion exchange and other methods, cost estimates were based on the cost procedures established by the Office of Saline Water (OSW) [12]. The costs were corrected to 1965 on the basis of the Engineering News-Record Construction Cost Index (ENRCCI) [13]. Certain OSW costs and provisions were modified as follows:

1. The construction period was assumed to be six months instead of two years which reduced the construction charges.
2. The facilities costs which could have been obtained from the OSW publication (viz. tanks, motors, and pumps) were obtained from other sources deemed more reliable for the small sizes involved.
3. Erection costs were obtained from the literature when possible. Otherwise, according to OSW procedure, erection was assumed to be 30 per cent of the installed price, except that erection costs of motors and the instrument shed were assumed to be 100 per cent.
4. The maintenance labor and supplies specified by the OSW were deemed to be inadequate for small ion exchange plants and were therefore doubled.

It is hoped that these modifications provide realistic costs and, at the same time, a reasonable comparison with other costs published by the OSW and others.

The costs of ion exchange facilities and installations were obtained from the work of Peak and David [14]. Costs of tanks and pumps were obtained from Page [15]. The costs of motors and concrete were obtained from Means [16]. The costs of some storage tanks, the sand filter,

chemicals, and freight rates were obtained from companies with offices in the San Francisco Bay area. All such costs were verified to ensure credibility. It was assumed that ion exchange facilities would be shipped 1000 miles by truck at a rate of \$3.00/cwt for 12-ton minimum loads and at a rate of \$3.60/cwt for LTL (less than truck loads).

The costs of resins in the small quantities needed for producing 0.1 mgd of product water were assumed to be: \$21.00/cu ft of Duolite C-25, \$27.60/cu ft of Amberlite 200, \$67.40/cu ft of Amberlite IRA-93 in its irreversibly swollen OH form. The assumed cost of chemicals transported 300 miles is given in Table VI.

TABLE VI
CHEMICAL PRICES FOR COST ESTIMATES

Chemical	Price, f.o.b.	Freight 300 miles	Total
Sulfuric acid, 66°Be 92.6% H ₂ SO ₄	\$24.65/ton acid	\$0.64/cwt	\$ 40.45/ton as H ₂ SO ₄
Ammonia, 30% aqueous	76.00/ton NH ₃	0.64/cwt	118.70/ton as NH ₃
Sodium hydroxide, 50% aqueous	56.00/ton of 76% Na ₂ O	0.64/cwt	81.60/ton as NaOH
Sodium carbonate	27.50/ton	0.56/cwt	38.70/ton as Na ₂ CO ₃

The utilization of chemicals, the flow rates, the cycle time, and the leakages were based upon the laboratory results obtained from pilot plant operations. It was assumed that:

1. Exchange efficiency is independent of feed concentration as indicated by efficiencies of 91.7 per cent for 770 mg/l TDS feed and, under nearly similar operating conditions, 93 per cent for 2000 mg/l TDS feed.

2. Leakage is a straight line function of feed concentration and anion leakage equals cation leakage.
3. The amount of rinse needed for all feeds is that used in the pilot plant operations.

Cost estimates are based on C-25 cation resin because less rinse, higher efficiencies, and lower leakages were obtained with its use.

The foregoing assumptions and discussion provide a rational basis for computing the operating characteristics for water containing 2500 mg/l or less of mineral constituents.

Optimum Level of Regeneration

To determine the optimum level of regeneration of cation resin the results of the cyclic investigations were converted to a feed containing 2000 mg/l TDS. The cost of cation exchange for each of the three levels of regeneration investigated was estimated assuming recovery and reuse of acid by storage in wood tanks for 20 hours followed by filtration to remove calcium sulfate precipitate.

The estimated costs shown in Figure 12 for a 0.1 mgd capacity plant indicate that 1.25 lb H_2SO_4 /cu ft is nearly the optimum level of regeneration. As Figure 7 indicates, the efficiency must decline rapidly in the region below 1.25 lb H_2SO_4 /cu ft because the water required for operation becomes a significant proportion of the product. Moreover, the leakage at 1.25 lb H_2SO_4 /cu ft was close to the tolerable limit. Hence it was concluded that lower levels of regeneration need not be investigated.

Reuse of Acid

If the acid regenerant for cation resin is not reused the net efficiency decreases from 84 per cent to 70 per cent for C-25 resin as shown in Table II. For a plant producing 0.1 mgd of water partially demineralized from 2000 ppm to 500 mg/l, the corresponding difference in the acid requirement is approximately 260 lb of commercial sulfuric acid per day which costs about five dollars. Assuming a life of 10 years and an interest rate of 4 per cent, this daily operating cost is approximately equal to a capital expenditure of about \$14,000 which corresponds to a purchase price (neglecting erection, engineering, contingencies, taxes, etc.) of about \$10,000. Unless the storage tanks, filters, and filter cake handling facilities can be installed at less cost, it is uneconomical to attempt to save acid by this means. Moreover, such facilities are not standard, so they would have to be specially designed, and they would be subject to the kind of difficulties that frequently afflict such custom-made facilities. Furthermore, there are added hidden costs such as increased labor, more complex piping and instrumentation, and longer regenerating time which requires larger reactors.

It seems unlikely that facilities for the storage of acid and the handling of the precipitate can be obtained at a sufficiently attractive price. Furthermore, such facilities would probably require considerable attention and increase the labor charges. Hence, the reuse of acid was considered to be uneconomical in a 0.1 mgd installation and the estimates for the cost of ion exchange were based upon regeneration with new acid only.

Estimated Cost of Ion Exchange

A schematic diagram of an ion exchange plant to produce 0.1 mgd of potable water is shown in Figure 13. The design is based upon the operating characteristics given in Tables III and V. The influent is assumed to be the synthetic water of Table I. The effluent contains only the dissolved minerals from the leakage which is about 388 mg/l. By blending the effluent with raw water to raise the TDS in the finished water to the 500 mg/l limit, the unit product cost can be reduced about 7 per cent. The automatic sand filter shown might not be required since the resin itself is an excellent filter. However, it would reduce the necessary frequency of backwashing the resin and would inhibit fouling due to turbidity. The resins were assumed to be Duolite C-25 (regenerated with new sulfuric acid only) and Amberlite IRA-93 (regenerated with ammonia). The estimated costs to construct and to operate this plant are given in the Appendix.

The costs to treat water of other salinities are shown in Figure 14. The operating characteristics were computed upon the basis of the assumptions listed on pages 25 and 26, and the cost estimates followed the procedure set forth in the Appendix. Both well water and synthetic water contained a high proportion of bicarbonate alkalinity (40 per cent of the anions on a me/l basis) and many natural waters contain much less bicarbonate which would increase the load upon the anion exchanger and require more ammonia. An approximate correction for the cost of ion exchange for water of a different bicarbonate content can be made by assuming that the ammonia required is directly proportional to the concentration of strong acid anions. The upper limit of cost is reached when the bicarbonate content is high enough to require a degasifier yet low

enough to require the maximum amount of ammonia. For the purpose of cost estimation, this condition can be approximated by assuming the water to contain no bicarbonate, but retaining, nevertheless, the degasifier. The slight modifications of reactor size would have scant effect upon the operating cost. The dashed lines in Figure 14 show the costs for water containing no bicarbonate.

The significance of the costs in Figure 14 can be appreciated by comparing them to the costs in Table VII, in which the only costs less than \$1.00/1000 gal are those for the Buckeye electrodialysis plant and estimated future costs for extremely large sea water conversion plants. Buckeye water is warm and contains mainly sodium chloride and is ideal for electrodialysis. On the other hand, Webster water is cold and contains calcium, iron, manganese, and sulfate and is poorly adapted to electrodialysis. For small plants treating between 0.05 and 0.25 mgd of moderately brackish water containing 2500 mg/l or less of dissolved solids, ion exchange appears to be competitive with electrodialysis and is much cheaper than distillation.

TABLE VII

COST OF DESALINIZATION

Location and Type	TDS, mg/l		Capacity mgd	Operating Cost \$/K gal
	Input	Output		
<u>Actual Costs of Sea Water Conversion</u>				
Aruba Island, Venezuela Multiple effect distillation [17]	35,000	20	3	1.75
Freeport, Texas Multiple effect LTV [17]	35,000	20	1	1.51
San Diego, California Multistage flash [17]	35,000	25	1	1.16
<u>Actual Costs of Brackish Water Conversion</u>				
Buckeye, Arizona Electrodialysis [18]	2,100 2,100	484 490	0.65 0.31	0.33 0.51
Webster, South Dakota Electrodialysis [19]	1,500	250	0.25	1.18
Coalinga, California Electrodialysis [20]	2,000	400	0.028	1.43
<u>Estimated Future Costs of Sea Water Conversion</u>				
Largest practical fossil fuel plant [21]	35,000	20	170	0.50
Nuclear powered [21]	35,000	20	170	0.45
Nuclear powered [21]	35,000	20	1300	0.26

CONCLUSIONS

The investigation has demonstrated that an ion exchange system, consisting of a strong acid resin, a degasifier, and a weak base resin, is an economically feasible process for producing drinking water from brackish sources containing up to 2500 mg/l of dissolved minerals. At salinities of 2000 mg/l the cost of partial demineralization by ion exchange is approximately \$0.89/1000 gallons in a plant large enough to process about 0.1 mgd. In comparison, the cost of electrodialysis in larger plants varies from \$0.33/1000 gallons to more than \$1.18/1000 gallons and these costs would be greater in smaller plants. The cost of electrodialysis is more strongly affected by temperature, ionic species, and ratio of average to peak demand than is ion exchange.

The cost of operation of a fully automatic 0.1 mgd plant to treat a moderately brackish water containing 2000 mg/l of dissolved minerals can be divided approximately as follows:

Amortization		Power	0.7%
Sand filter, pump, motor . . .	3.4%	Acid	30.2
Cation reactor, pump, etc. . .	4.8	Ammonia	12.1
Degasifier and tank	1.8	Labor	17.1
Anion reactor, pump, etc. . . .	4.8	Maintenance	5.0
Chemical storage, pumps	4.6	Overhead, Taxes,	
Instruments, shed	<u>4.2</u>	Insurance	<u>11.3</u>
Total	23.6	Total	100.0

The most effective ways to reduce the cost are to save chemicals, reduce labor, and reduce the size of the facilities. Chemicals can be used more efficiently by regenerating the cation resin at a level of 1.25 lb H_2SO_4 /cu ft, or perhaps slightly less. The reuse of the spent acid may effect additional savings if the necessary facilities can be installed at a

sufficiently low price. Unfortunately, the storage, precipitation, and filtration required for acid recovery are likely to be too expensive in small installations and to require substantial labor. The cost of ammonia might be reduced by recovery with lime, but unless the process is fully automated, labor charges are likely to cancel any savings.

Advantageously, the manufacture of ion exchange equipment is competitive and the facilities are standardized which tends to keep the purchase price low. The process is well adapted to automatic control, and with safety devices, the plants require no attendance beyond occasional inspection. The estimated labor costs provide for daily inspection and a generous amount of time for servicing instruments to ensure trouble free operation.

The power requirements are small and the size and cost of the facilities might be reduced slightly at flow rates higher than those employed in these investigations. However, the operations reported herein must be close to the optimum.

Since the efficiency is relatively independent of salinity, the procedure proposed herein provides a rational means for designing the demineralization cycle. However, if the relative concentrations of the ionic species differ greatly from those of the water investigated there may be an effect on the efficiency and operation. For example, if there is little potassium and sodium, the efficiency may be decreased and the precipitation of calcium may become troublesome.

Unfortunately, the economic limitations of ion exchange cannot be unequivocally stated, but will depend upon the cost of other deionization methods and the cost of an alternative, even if distant, water supply. Nevertheless, it has been shown that there are conditions under

which ion exchange compares favorably with other means of deionization. Each situation requires individual study to determine whether ion exchange is the most economical alternate.

ACKNOWLEDGMENTS

The investigation was carried out under the provisions of Grant WP 00544 by the National Institutes of Health, USPHS, to the University of California at Berkeley. Charles Smedley, Neil Nininger, and James Larson assisted in the laboratory work during various periods. The helpfulness of Professors Theodore Vermeulen and J. F. Thomas of the University of California, and Dr. Irving Abrams of Diamond Alkali Company is gratefully acknowledged. Exchange resins were supplied by Rohm and Haas Company and by Diamond Alkali Company.

APPENDIX

COST TO DEMINERALIZE 2000 mg/l TDS FEEDWATER IN 0.1 mgd ION EXCHANGE PLANT

(Based on flow diagram in Figure 13)

Capital Investment

Sand filter. Tank, coagulant mixer, solution tank and pump, automatic controls, assembled*, freight, plus erection at 30%	\$ 5,322
Clearwell. Excavation, forms, reinforced concrete [16], and pipe fittings [15], plus erection at 30%	1,447
Cation reactor. Foundation [16], rubber-lined tank [14], resin, freight, 10% incidentals, plus erection at 30% .	10,357
Degasifier. Foundation [16], redwood tower* and tank [15], 1/3 hp blower and motor*, freight, installation [15], plus erection at 30%	4,429
Anion reactor. Foundation [16], rubber-lined tank [14], resin, freight, 10% incidentals, plus erection at 30% .	10,584
Acid storage. Foundation [16], 4000 gal placite-lined vertical tank with flanges, manhole, valves and miscellaneous hardware, installed*, freight, plus erection [15] at 51%	5,075
Ammonia storage. Foundation [16], 7000 gal vertical tank with manhole and flanges*, valves and miscellaneous hardware, installed*, freight, plus erection [15] at 41%	4,160

*Bid price, local company.

Instruments. Automatic cation column controls [14], anion column controls [14], freight, 10% incidentals, plus erection at 30%	\$ 9,306
Pumps. 300 gpm filter pump [15], 120 gpm cation column stainless steel pump [15], 90 gpm anion column stainless steel pump [15], acid pump and mixer*, ammonia pump and mixer*, installation, plus erection at 30% . .	6,370
Motors [16]. 5 hp filter pump motor, 3 hp cation column pump motor, 3 hp anion column pump motor, 1/3 hp blower motor, plus erection at 100%	690
Instrument shed. Concrete floor [16], 8 x 12 x 8 ft galvanized iron shed*, freight, plus erection* at 100%	<u>1,242</u>
Total essential plant cost	\$58,982
Other plant costs	
Raw water supply** at \$7.17/1000 gpd	829
Product storage at \$14.33/1000 gpd	1,433
Contingencies at 10% of above	6,124
Engineering at 10% of above	6,737
Interest† during construction at 1%	741
Plant site**	<u>430</u>
Total plant investment (TPI)	\$75,276
Working capital, 60 days production at operating cost . . .	5,760
Total cost per gpd	\$ 0.81

*Bid price, local company.

**OSW [12] cost corrected by ENRCCI [13] = 975/680.

†OSW [12] assumes 4% for two-year construction.

Operating Costs

Electric power at 7 mills and 100 kwh	\$ 0.70
Sulfuric acid	28.90
Ammonia	11.62
Resin replacement at 5%/year	0.33
Supplies and maintenance at 0.003%* TPI	2.26
Operating labor, 8 hr/wk engineer at \$5.25/hr plus 16 hr/mo instrument service at \$15/hr plus maintenance labor at 0.003%* TPI	16.26
Payroll extras at 15% labor	2.44
General overhead at 30% payroll	5.61
Amortization. 20 yr life for 31% TPI, 15 yr life for 25.5% TPI, 10 yr life for 39.5% TPI, and 7 yr life for 4% TPI	22.65
Taxes and insurance at 0.006% TPI	4.51
Interest on working capital, 4% of 60 days production plus 10% (6 days) inoperative time	0.69
Total operating cost per stream day	95.97
Cost per 1000 gal water at 388 mg/l TDS	0.96
Cost per 1000 gal blended water (107,500 gpd) at 500 mg/l TDS	0.89

*OSW specifies 0.0015%.

REFERENCES

1. Bachmann, F. "The economics of de-ionized water supplies." J. AWWA 36:876-85, August 1944.
2. Spiegler, K., W. Juda, and M. Carron. "Counterflow regeneration of cation exchanger in partial demineralization of brackish waters." J. AWWA 44:80-88, January 1952.
3. Yoseph, R. "Demineralization of brackish water for potable supply at Matagorda Island." J. AWWA 48:579-84, May 1956.
4. Kunin, R. and B. Vassiliou. "New ion exchange resin techniques for deionizing brackish waters." Sonderdruck aus Dechema-Monographien, Band 47:735-50, 1962.
5. Kunin, R. and B. Vassiliou. "New deionization techniques based upon weak electrolyte ion exchange resins." I&EC Process Design and Development 3(4):404-09, October 1964.
6. Sturla, P. "Pilot plant studies of the Kunin process." Paper, presented at 25th meeting, Internat. Water Conf., Engineers' Soc. of Western Pa., Pittsburgh, September 1965.
7. The SUL-biSUL Process, a Method for Demineralizing Brackish Water for Potable and Other Uses. Bull. Z-18, copyright 1965 by Nalco Chemical Company, Chicago, Illinois.
8. Standard Methods for the Examination of Water and Wastewater, 11th Ed., APHA, New York, 1960.
9. Jenkins, D. and L. Medsker. "Brucine method for the determination of nitrate in ocean, estuarine, and fresh waters." Anal. Chem. 36: 610-12, March 1964.
10. Kahler, F. and A. Reents. "Water demineralization costs." Chem. Eng. 64:206-10, January 1957.
11. Amberlite Ion Exchange Resins Laboratory Guide. Rohm and Haas Co., March 1964.
12. U.S. Dept. of the Interior, Office of Saline Water. A Standardized Procedure for Estimating Costs of Saline Water Conversion. March 1956.
13. Engineering News-Record 174:73-74, March 18, 1965.
14. Peak, R. and M. David. "Costs of cation exchange equipment." Chem. Eng. Progress 53:37J-40J, January 1957.

REFERENCES (continued)

15. Page, J. Estimator's Manual of Equipment and Installation Costs.
Houston: Gulf Publishing Co., 1963.
16. Means, R. Building Construction Cost Data 1965. Duxbury, Mass.:
Means Co., 1965.
17. Schneller, F. "Where we stand on saline water conversion." Civ. Eng.
31:38-43, December 1961.
18. Kirkham, T. "Report on design, installation, start-up and operation
of Ionics, 650,000 gpd electrodialysis plant, Buckeye, Arizona."
Presented at Div. Water and Waste Chemistry, Amer. Chem. Soc.,
Los Angeles, April 1963.
19. Office of Saline Water. Saline Water Conversion Report for 1963.
Supt. of Documents, Washington, D.C., 1963.
20. Cary, E., H. Ongerth, and R. Phelps. "Domestic water supply
demineralization at Coalinga." J. AWWA 52:585-93, May 1960.
21. Revelle, R. Large Nuclear Powered Sea Water Distillation Plants.
Office of Science and Technology, Interagency Task Group, Supt.
of Documents, Washington, D.C., March 1964.

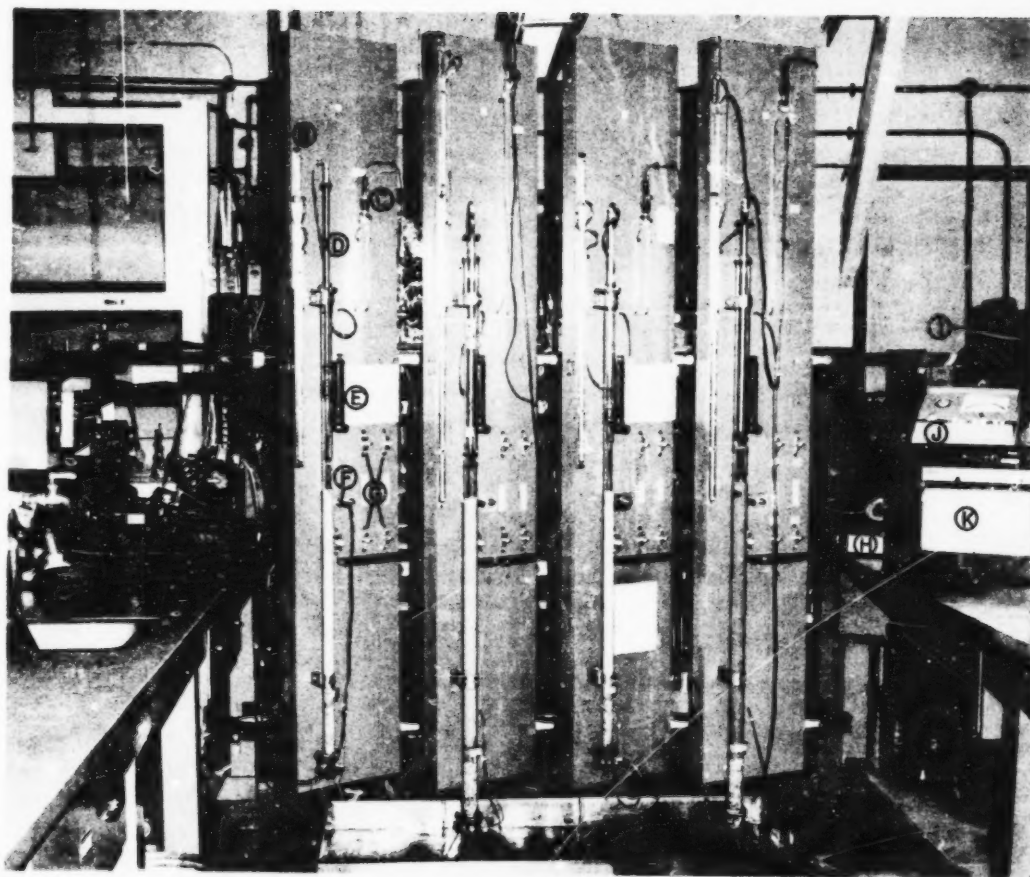


FIGURE 1. ION EXCHANGE PILOT PLANTS

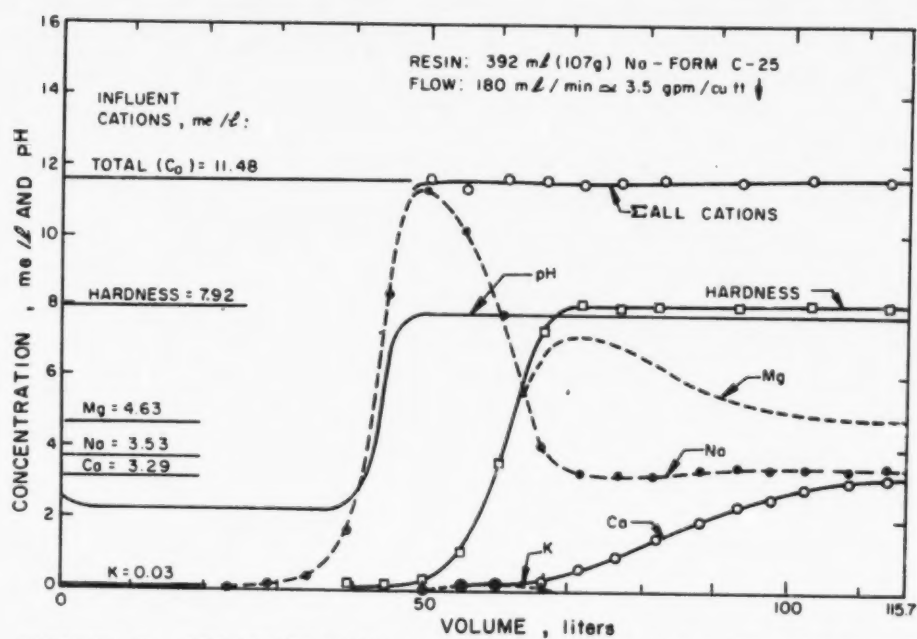


FIGURE 2. CONCENTRATION HISTORIES FOR C-25 RESIN EXHAUSTED TO EQUILIBRIUM WITH WELL WATER

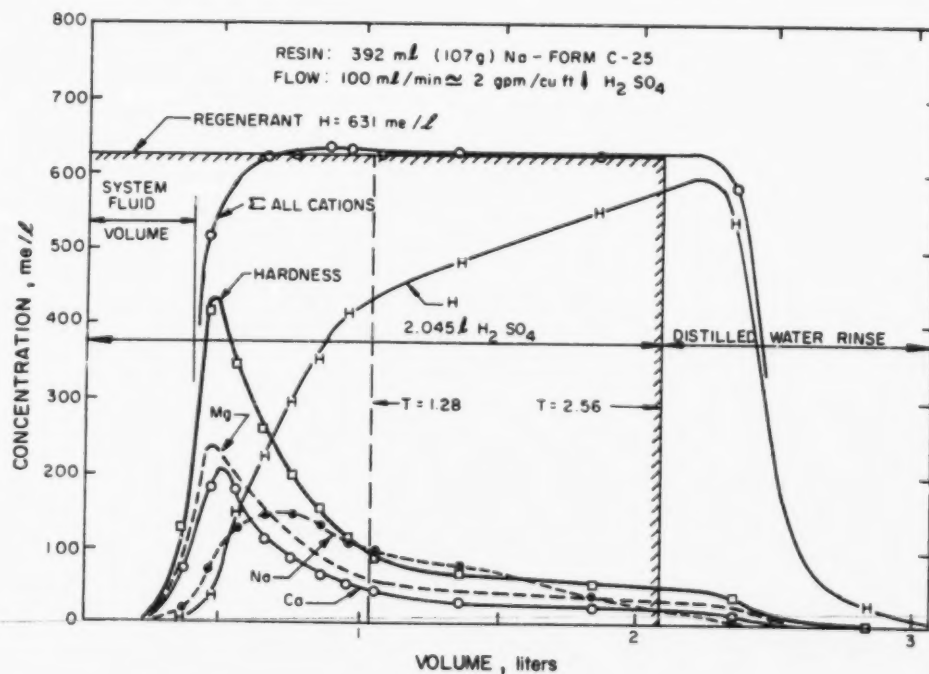


FIGURE 3. TYPICAL CATION REGENERATION CONCENTRATION HISTORIES

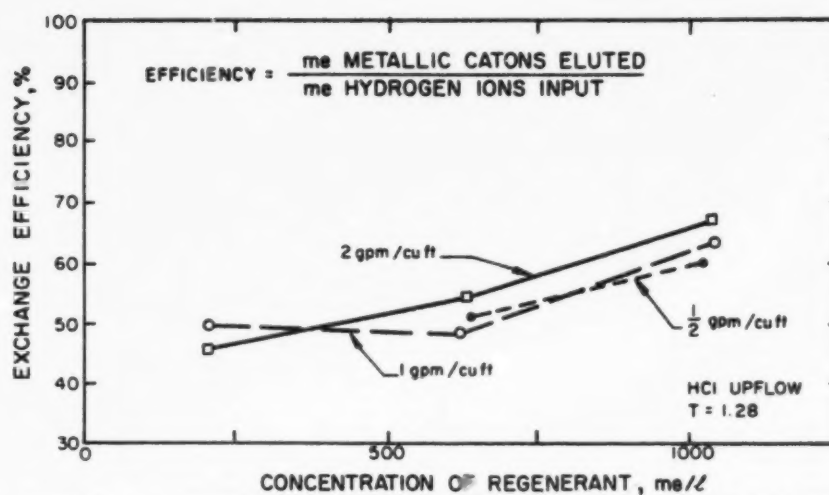


FIGURE 4. EFFECT OF CONCENTRATION OF REGENERANT ON EXCHANGE EFFICIENCY

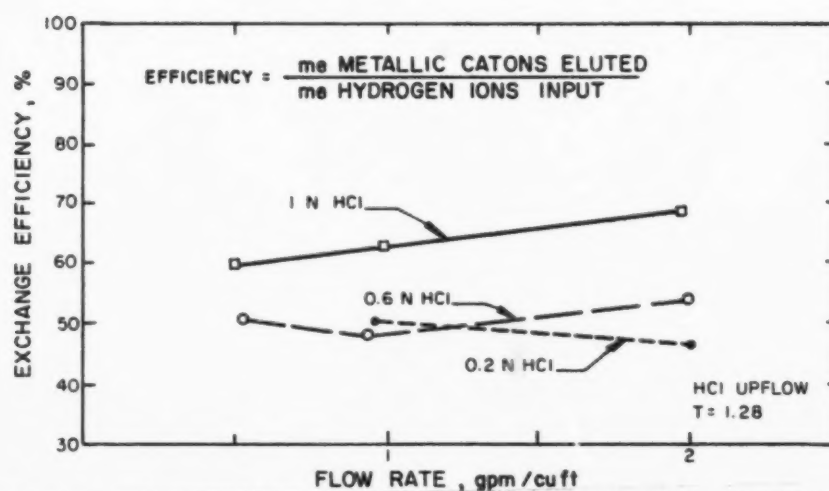


FIGURE 5. EFFECT OF FLOW RATE OF REGENERANT ON EXCHANGE EFFICIENCY

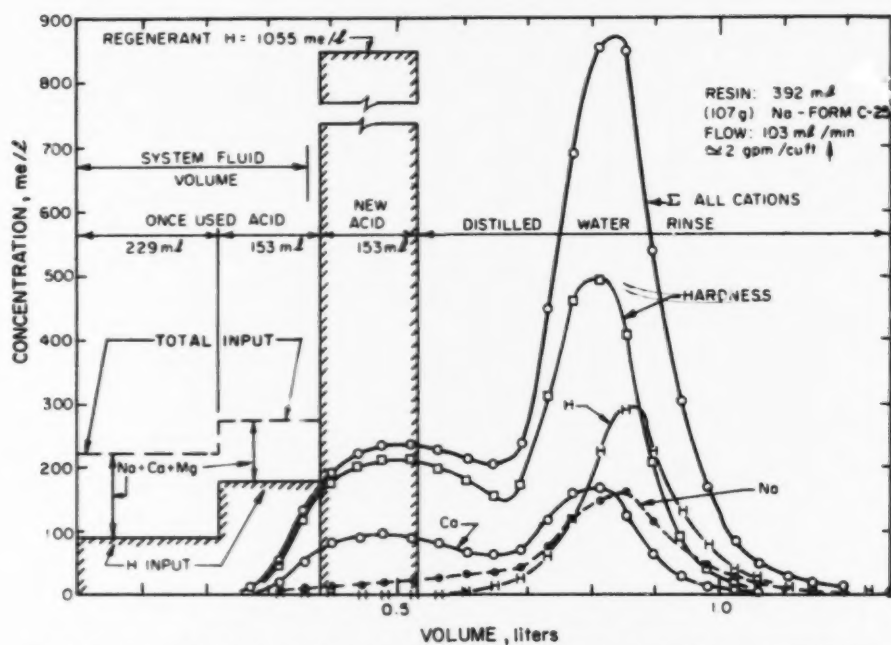


FIGURE 6. TYPICAL CATION REGENERATION CONCENTRATION HISTORIES IN CYCLIC OPERATIONS WITH ACID REUSE

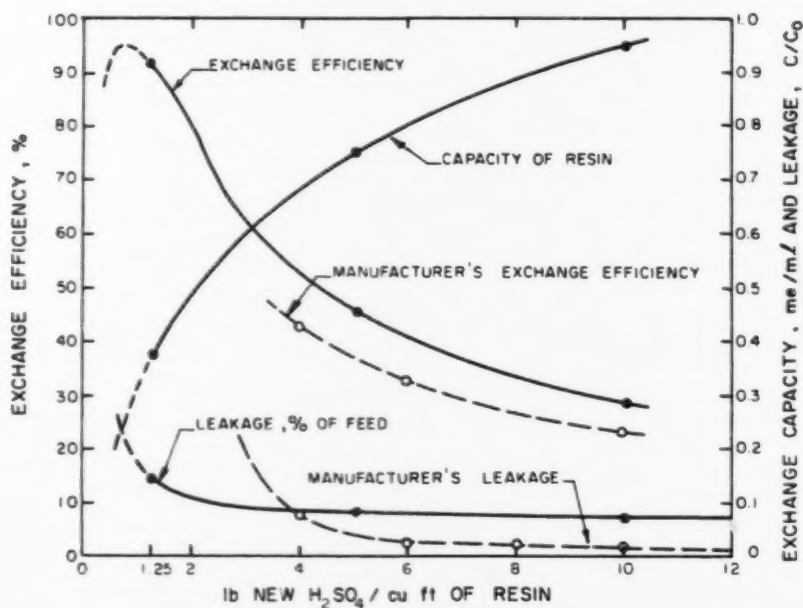


FIGURE 7. RELATION OF CATION EXCHANGE EFFICIENCY TO REGENERATION LEVEL WITH ACID REUSE (C-25 RESIN, 5% H₂SO₄ COUNTERCURRENT AT 2 gpm / cu ft)

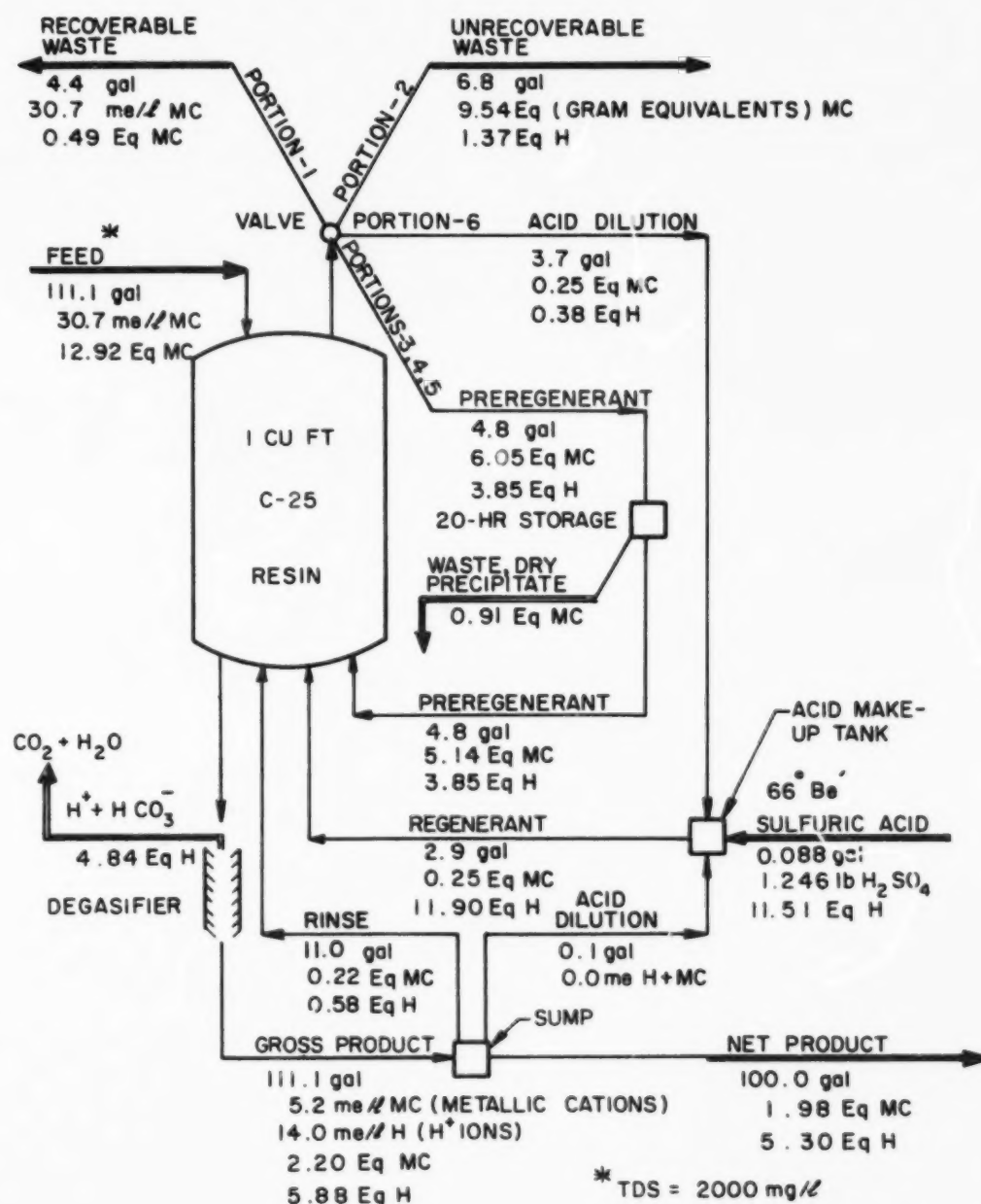


FIGURE 8. MATERIAL BALANCE FOR AN OPTIMUM C-25 CATION EXHAUSTION - REGENERATION CYCLE

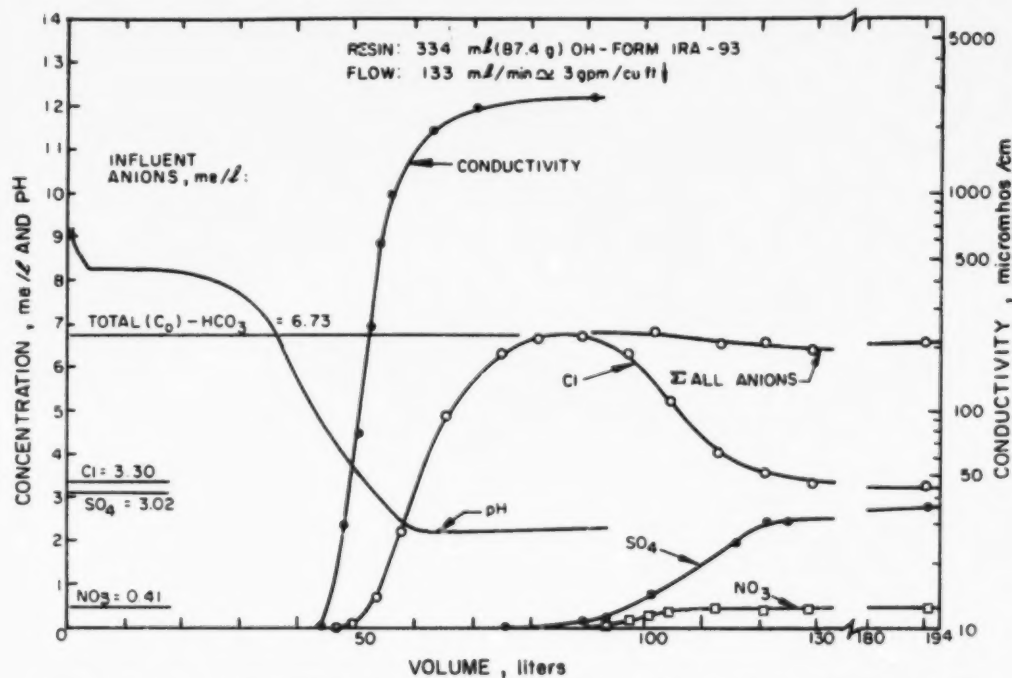


FIGURE 9. CONCENTRATION HISTORIES FOR IRA-93 RESIN EXHAUSTED TO EQUILIBRIUM WITH DECATIONIZED WELL WATER

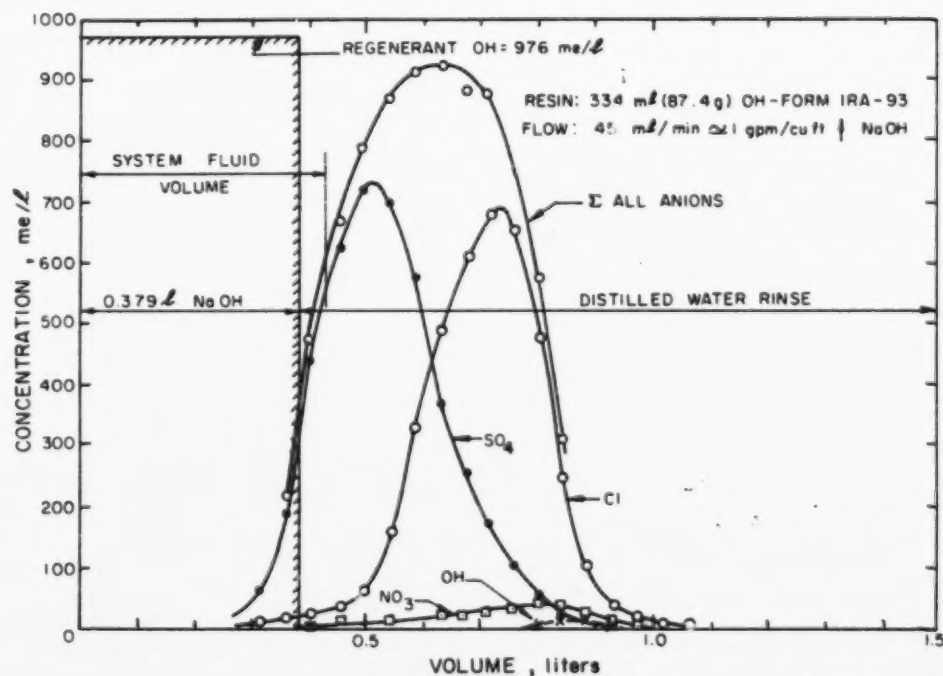


FIGURE 10. TYPICAL ANION REGENERATION CONCENTRATION HISTORIES

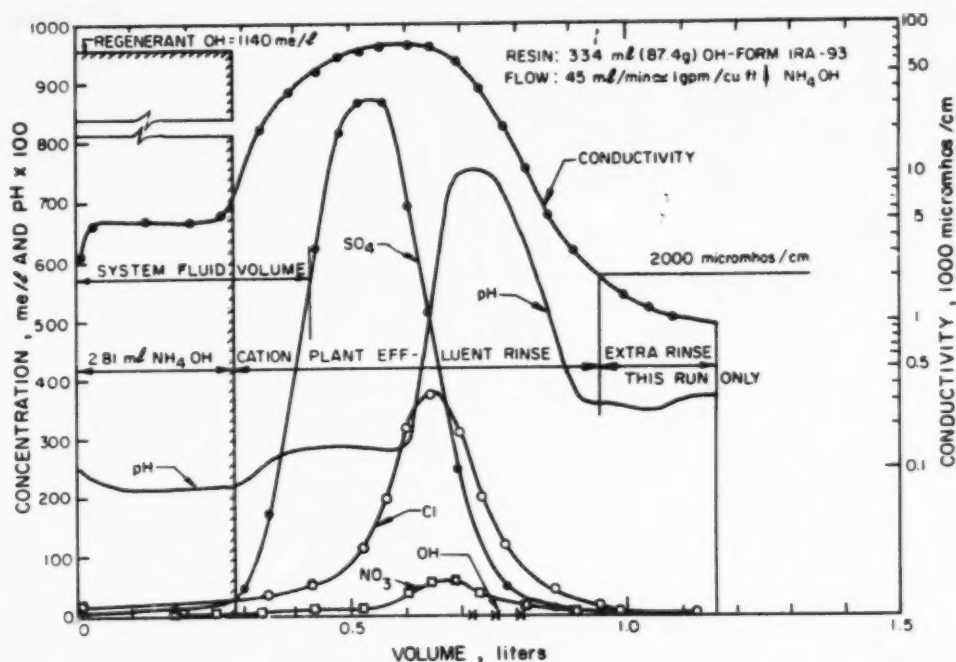


FIGURE 11. TYPICAL ANION REGENERATION CONCENTRATION HISTORIES IN PILOT PLANT OPERATIONS

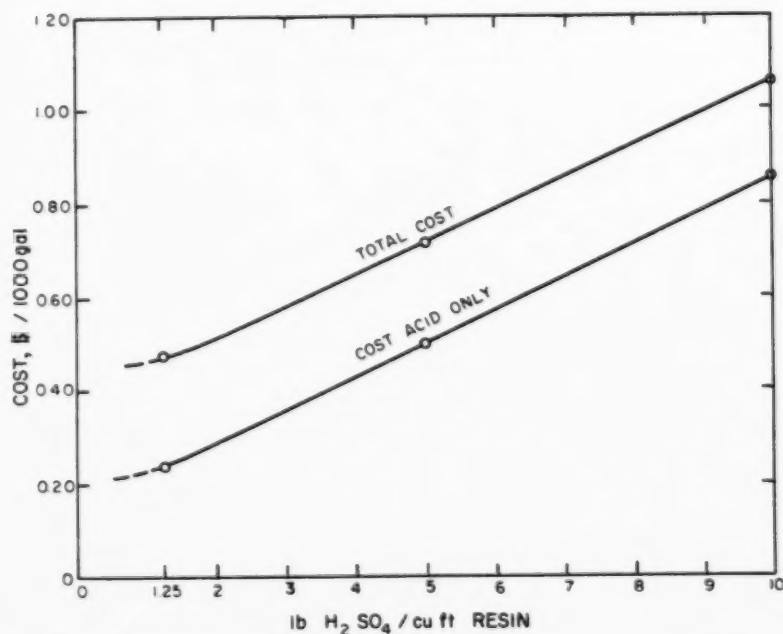


FIGURE 12. ESTIMATED COST OF CATION EXCHANGE FOR 2000 mg/l TDS FEED, 500 mg/l TDS EFFLUENT WITH ACID REUSE

RECEIVED
NATIONAL TECHNICAL
JAN 15 1973
INFORMATION SERVICE

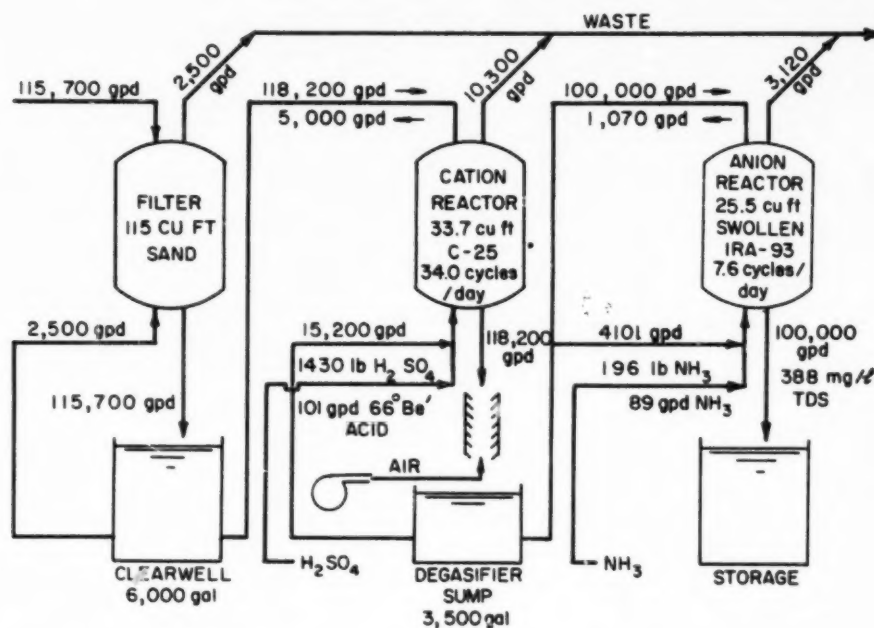


FIGURE 13. SCHEMATIC FLOW DIAGRAM FOR 0.1 mgd ION EXCHANGE PLANT TO TREAT 2,000 mg/l TDS FEED

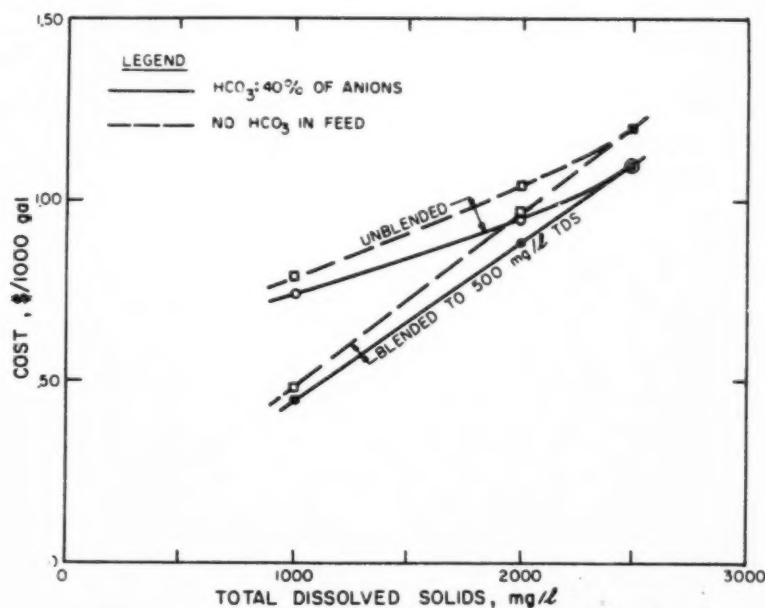


FIGURE 14. COST OF ION EXCHANGE FOR PARTIAL DEMINERALIZATION OF SALINE WATER